

MODELLING OF INDIRECT STEAM GASIFICATION IN CIRCULATING FLUIDIZED BED REACTORS

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Abstract – The indirect steam gasification in circulating fluidized bed reactors was studied by modelling. The object of study was a coupled 12 MWth gasifier-combustor system, which was fired by woody biomass. The heat for the steam-blown gasifier was produced in the air-blown combustor and transported by circulating solids between the interconnected reactors. The system was modelled by a semi-empirical three-dimensional model, which simulated the fluid dynamics, reactions, and heat transfer in the coupled process. The studied cases included different temperature levels, which were controlled by the amount of additional fuel feed to the combustor. The model concept can be later applied to study sorbent enhanced gasification, which is a promising method for sustainable production of transport fuels to substitute fossil based fuels.

INTRODUCTION

Gasification of biomass is a method to produce renewable, carbon neutral energy, which, unlike wind and solar, is not dependent on weather conditions. The gasification produces syngas, which contains high share of combustible gas components: carbon monoxide, hydrogen, and hydrocarbons. The syngas can be combusted for producing heat and electricity, but it can also be converted to different products, such as synthetic natural gas (SNG), Fischer–Tropsch diesel, dimethyl ether, and methanol.

The gasification technologies are categorized to direct (autothermal) gasification and indirect (allothermal) gasification. In direct gasification, the fuel is partially oxidized to generate heat to sustain the gasification process. This can be accomplished by air-blown or steam/oxygen-blown methods. The disadvantage of air-blown method is the high amount of nitrogen in the syngas, which decreases the heat value of the gas. This is avoided in steam/oxygen-blown method, but this is expensive due to need for air separation unit to produce oxygen. In indirect gasification, the heat for gasification is imported from outside the gasifier reactor. This can be accomplished by heat exchangers, but this is thermally inefficient and technologically difficult due to high temperature requirements. Another method is to apply circulating hot solids as heat carriers, which is the object of this study.

Fig. 1 presents the basic principle of indirect steam gasification of biomass. The total process consists of two interconnected reactors: a gasifier and a combustor. The hot bed material, which consists of fuel ash, make-up sand, and sorbent, is transported from air-blown combustor to gasifier, where it provides heat for the gasification reactions. The syngas from the steam-blown gasifier is free from atmospheric nitrogen and mainly consists of combustible gases (CO , H_2 , C_xH_y), water vapour, and carbon dioxide. The cold bed material from the gasifier is transported back to combustor to be re-heated. The remaining unreacted char in the circulating solids is combusted in the combustor. Depending on the desired conditions, this can be the only fuel source for the combustor, or additional fuel can be fed to combustor to increase the temperature level of the process.

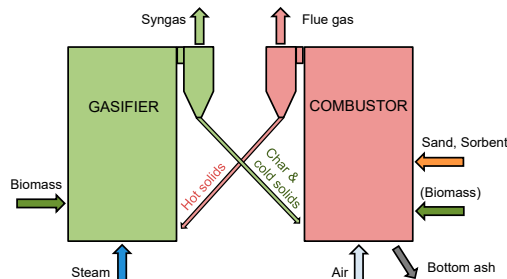


Fig. 1. Basic principle of indirect steam gasification of biomass.

Various concepts have been proposed for indirect steam gasification by using fluidized bed technology, but the basic principle is always the same as described above. In the FERCO SilvaGas™ process, both reactors operate in circulating fluidized bed (CFB) mode (Paisley et al., 2001). In the 8 MWth dual bed system in Güssing, the gasifier is operated in bubbling bed mode and the combustor in CFB mode (Pröll et al., 2007). In the MILENA process, the gasifier is operated in CFB mode and it is surrounded by a bubbling bed combustor (van der Meijden et al., 2011). In sorbent enhanced gasification (SEG), lime is applied to capture carbon dioxide in the gasifier and increase the share of hydrogen in the syngas. This method is investigated in a Horizon 2020 project FLEDGED (FLEXible Dimethyl ether production from biomass Gasification with sorption-enhanced processes).

The coupled reactors are complicated processes due to various interconnected phenomena. Thus, efficient and valid modelling tools are needed to support the development of new process concepts and scale-up. This paper presents a modelling method, which is suitable for simulating interconnected fluidized bed processes. This was applied to study the effect of different gasification temperatures on the syngas composition, heat value of gas, and cold gas efficiency. The object of study was a coupled gasifier-combustor system, in which both reactors were operated in CFB mode. The temperature level was adjusted by changing the fuel feed to combustor while keeping the fuel feed to gasifier at a constant value.

MODELLING METHOD

The coupled CFB reactors were simulated by a semi-empirical, steady-state engineering model (Hyppänen et al., 1991; Myöhänen and Hyppänen, 2011). The model combines fundamental balance equations with empirical correlation models and solves the process variables in a three-dimensional domain, which is discretized by a Cartesian structural mesh. The model solves fluid dynamics of gas and solids, combustion and gasification reactions, limestone reactions, attrition of solids, and heat transfer within the suspension and to heat transfer walls. The validation of the model is based on measurements ranging from pilot-scale to commercial-scale combustor and gasifier units (Koski et al., 2012; Lyytikäinen et al., 2014). Fig. 2 illustrates the modelled reaction paths for fuel conversion in a CFB.

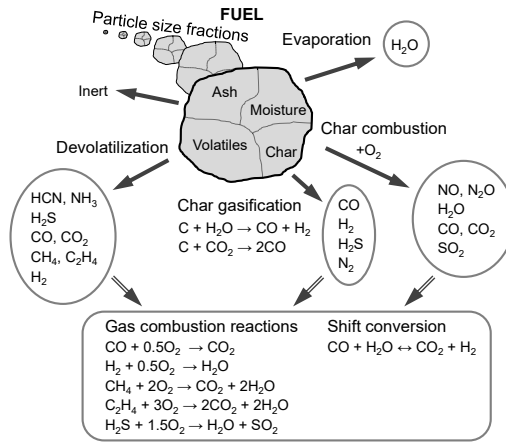
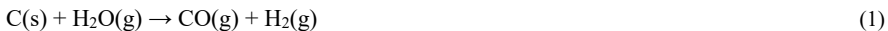


Fig. 2. Modelled reaction paths.

In the 3D model frame, the gasifier and combustor were directly coupled, i.e. the circulating solids separated in the cyclone of one reactor were applied as the incoming circulating material for the other reactor. The following presents the modelled main reactions to enable understanding the findings, which have been reported in the later sections of this work. In gasification conditions, the main heterogeneous reactions of fuel are evaporation of moisture, devolatilization, and gasification of char. In the 3D model, the evaporation and devolatilization rates and the composition of devolatilized gases are determined by empirical correlations (Myöhänen and Hyppänen, 2011). The modelled heterogeneous gasification reactions include water-gas reaction and Boudouard reaction:



The reaction rates are defined by the following equations, which are simplified forms of the expressions used by Petersen and Werther (2005):

$$r'''_{watg,i} = 235.3 C_{C,i} C_{H_2O} \exp\left(\frac{-15\,500}{T}\right) \quad (3)$$

$$r'''_{boud,i} = 7.696 \cdot 10^6 C_{C,i} C_{CO_2} \exp\left(\frac{-30\,600}{T}\right) \quad (4)$$

Shift conversion is an important reaction in gasification conditions. This is a reversible reaction, which affects the distribution of main gas components in the syngas:



In steam-blown gasification, the concentration of water vapour is high, thus, the direction of the reaction is usually towards right. The reaction rate is defined by Eq. (6), which is based on literature (Biba et al., 1978; Yoon et al., 1978), but corrected by a multiplier 0.1 according to Petersen and Werther (2005).

$$r'''_{shift} = 0.278 \exp\left(\frac{-1515.46}{T}\right) \left[C_{CO} C_{H_2O} - \frac{C_{CO_2} C_{H_2}}{0.0265 \exp\left(\frac{3956}{T}\right)} \right] \quad (6)$$

The current model does not simulate the formation of tars. The tars are complex heavy hydrocarbons, e.g. levoglucosan and naphthalene, which are formed during gasification and can cause fouling of the downstream equipment. In steam-blown gasification, the measured amount of tar in syngas has been in the order of 2...5 g/m³n,dry (Pfeiffer et al., 2009; Specht, 2009). This is equivalent to less than 0.1 mol-% share in the syngas, thus, it has relatively small impact on the syngas composition.

MODEL SETUP

Fig. 3a presents the modelled reactor system. The diameters of the gasifier and combustor were 1.6 m and 1.4 m respectively. The height of the reactors was 15 m. The solids separated by the cyclone of the gasifier were returned to combustor at height 0.3 m. The solids from combustor were returned to gasifier at height 0.5 m. The gasifier was fluidized by steam through the bottom of the reactor. Secondary steam flow was introduced at height 1.9 m together with the fuel feed. The combustor was fluidized by primary air through the grid. Secondary air was fed at height 1.1 m together with make-up sand and limestone. Additional fuel was fed to combustor at height 0.5 m.

Fig. 3b presents the model mesh for gasifier (left) and combustor (right). The dimensions of hexahedral calculation cells were 0.145...0.200 m. The number of cells was about 6300 for gasifier and about 3800 for combustor. With a coarse mesh, the calculation with the semi-empirical model is fast. The calculation time for all cases was in the order of few hours with a normal laptop computer. Some additional time was spent for adjusting the air flow rates so that the excess oxygen (in dry basis) was the same in all cases.

The applied fuel was woody biomass (Table 1). Additional solid feeds included inert make-up sand and limestone, which was set to be 100% CaCO₃. The boundary conditions are given in Table 2. The limestone feed was small, thus, the calcination and carbonation reactions were not affecting much on the gas compositions. The fuel feed and the air flow feed to combustor were the only variable parameters. Four cases were calculated with different fuel feed to combustor to achieve different temperature levels. The air flow rate to combustor was adjusted so that the oxygen concentration after the combustor was 3.94 vol-%,dry. The thermal boundary of the walls was set by specifying the walls as refractory lined with thickness of 0.1 m, thermal conductivity 0.5 W/mK, and outer shell temperature 100°C, i.e. assuming that the reactors are well insulated, but with a small heat loss through all walls.



Fig. 3 a) Reactor system b) Mesh.

Table 1. Fuel properties.

Proximate analysis, as fired (w-%)		Ultimate analysis, dry, ash-free (w-%)	
Fixed carbon	11.1	Carbon	51.0
Volatiles	61.7	Hydrogen	6.1
Moisture	25.0	Nitrogen	0.5
Ash	2.2	Sulphur	0.1
		Oxygen	42.3
Heat value (MJ/kg)			
HHV, dry solids	20.51		
LHV, as fired	13.76		

Table 2. Boundary conditions.

Parameter	Units	Case A01	Case A02	Case A03	Case A04
Steam flow to gasifier	(kg/s)	0.45
Primary steam ratio	(%)	40
Steam temperature	(°C)	180
Air flow to combustor	(kg/s)	1.84	2.06	1.62	1.38
Primary air ratio	(%)	50
Air temperature	(°C)	280
Fuel feed to gasifier	(kg/s)	0.9
Fuel feed to combustor	(kg/s)	0.10	0.15	0.05	0.00
Sand feed to combustor	(kg/s)	0.05
Limestone feed to combustor	(kg/s)	0.01
Solid feed temperatures	(°C)	30

MODEL RESULTS

Fig. 4 presents the modelled temperature and concentrations of oxygen, carbon monoxide and hydrogen at outer surfaces of the reactors of basic case (0.1 kg/s fuel feed to combustor). Fig. 5 presents the same data for the centre-planes of the reactors. In each image, the gasifier is located on left and combustor on right. The profiles were looking fairly similar in all cases, thus, only the basic case is presented as contour plots while line graphs are used for comparing the results of different cases.

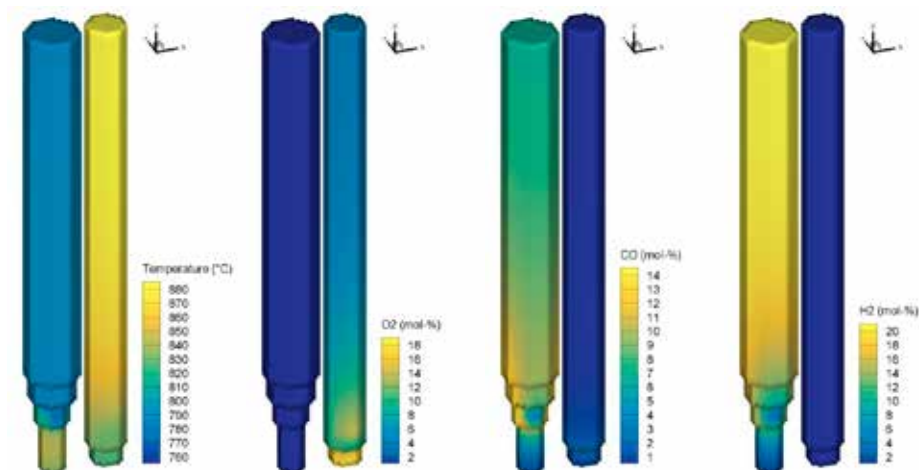


Fig. 4. Modelled 3D fields of basic case at outer surface of reactors.

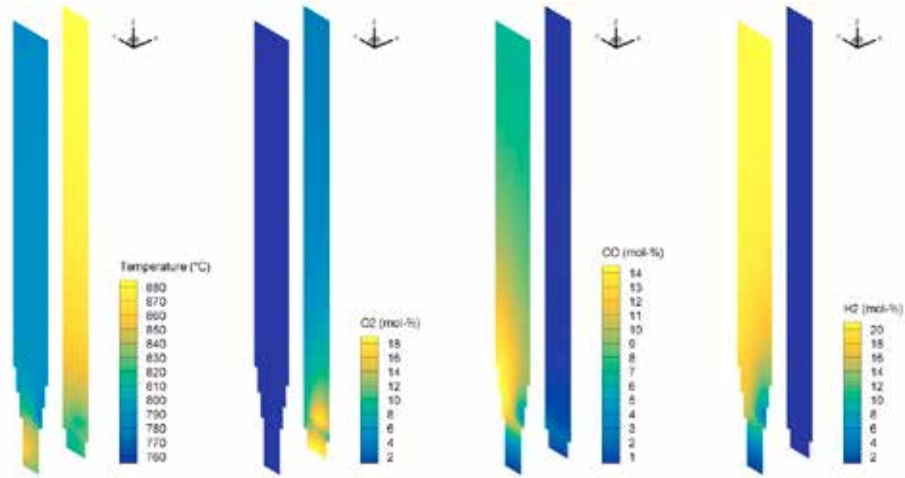


Fig. 5. Modelled 3D fields of basic case at centre-plane of reactors.

In the combustor, most of the fuel burns at the lower section of the furnace and consumes the oxygen, which originates from the primary and secondary air feeds. Some local CO and H₂ is found in the combustor near the fuel inlet, but the concentrations are very small. The combustion reactions heat up the circulating solids in the combustor. The hot solids from the combustor are transferred to lower section of the gasifier, where a local higher temperature zone is formed. At the location of the fuel feed to gasifier, the temperature decreases rapidly, mostly due to evaporation of fuel moisture. In the gasifier, the combustible gases, such as CO and H₂, originate from devolatilization and water-gas reaction. In the upper section of the gasifier, the CO content decreases and the H₂ content increases due to shift conversion. Based on the model results, the lateral mixing of solid and gaseous species needs to be considered near the fuel and air inlets and return legs, but at the upper sections, the lateral profiles are fairly uniform.

Fig. 6 presents the vertical gas concentration profiles of the gasifier for the basic case. The composition of combustible gases is mainly affected by devolatilization, water-gas reaction, and shift conversion, which are shown in Fig. 7. Because steam is used as fluidization gas, the concentration of water vapour is nearly 100% at the bottom of the gasifier. A local peak in the H₂O-profile can be seen at the level of the fuel feed, which is due to evaporation of fuel moisture. In the lower section, the share of H₂O decreases rapidly due to dilution by other gas sources: devolatilization and gasification of char. With biomass, the share of volatiles is high, thus, the source of CO from gasification of char (water-gas reaction) is much smaller than the source of CO from volatiles. In this case, the production of CO and H₂ from volatiles was about identical in molar basis. At the upper section of the gasifier, the H₂ and CO₂ concentrations are increasing while H₂O and CO concentrations are decreasing. This is due to shift conversion, which is highest just above the fuel feed and then continues throughout the gasifier. The shift conversion rate decreases towards the top of the gasifier due to decreasing share of reactants (H₂O and CO).

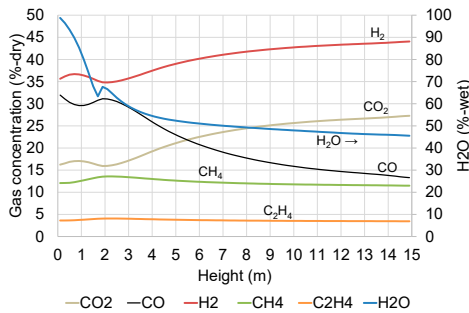


Fig. 6. Gas concentration profiles for basic case.

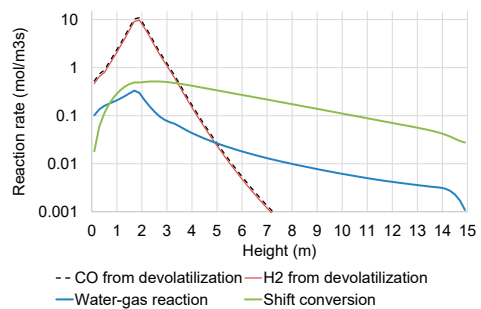


Fig. 7. Main reaction rate profiles for basic case.

Fig. 8 presents the vertical temperature profiles. The highest temperature in the gasifier is near the entry point of hot solids from the combustor. Above that, the temperature decreases due to endothermic reactions. At the upper part, the temperature is almost constant, i.e. the exothermic shift conversion is balanced by the heat loss through the walls and endothermic gasification reactions. In the combustor, the temperature is increasing as a function of height due to combustion reactions, which continue throughout the furnace with elutriative wood. The temperature difference between the outlets of combustor and gasifier was 92...103°C (lower difference with higher load). The vertical gas velocities are shown in Fig. 9. In the combustor, the average velocity increased from 3.8 to 6.8 m/s as the fuel input to combustor increased from zero to 0.15 kg/s. In the gasifier, the velocity was affected mainly by temperature and the average velocity varied from 2.5 to 3.0 m/s.

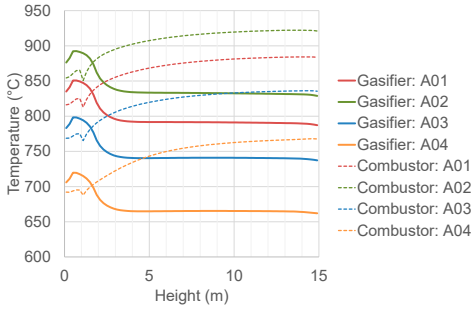


Fig. 8. Temperature profiles.

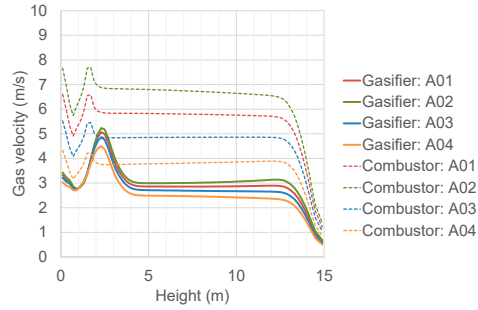


Fig. 9. Vertical gas velocity profiles.

The gas concentration of the syngas is presented in Fig. 10. Because the amount of fuel feed to the gasifier is constant, and thus, the amount of gases from devolatilization is not changing, the gas concentration is mostly affected by changes in the shift conversion and the water-gas reaction (Fig. 11). The water-gas reaction rate by Eq. (3) increases as a function of temperature. On the other hand, with the modelled gas compositions, a higher temperature results to smaller shift conversion as calculated by Eq. (6). A higher water-gas reaction increases the production of CO and H₂ from char. At the same time, the decreasing shift conversion reduces the conversion of CO and H₂O to CO₂ and H₂. As seen in Fig. 11, the changes in the two reactions are almost the same, but in opposite direction. As a consequence, the H₂ concentration is not much changing as a function of temperature, because the two reactions are compensating each other. The CO₂ level is decreasing due to decreasing shift conversion as the temperature increases. The CO level is increasing due to increasing water-gas reaction and decreasing shift conversion.

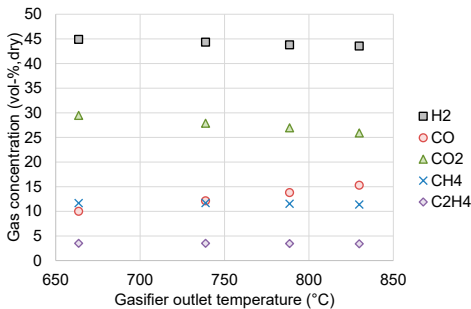


Fig. 10. Syngas concentration.

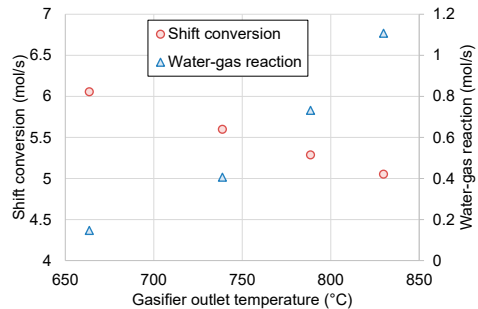


Fig. 11. Shift conversion and water-gas reaction.

Fig. 12 presents the lower heat value of syngas and the cold gas efficiency in different cases. The cold gas efficiency (sometimes called chemical efficiency) is defined as the chemical heat of the produced syngas divided by the total fuel power fed to gasifier and combustor:

$$\eta_{CG} = \frac{q_{v,gas} Q_{gas,LHV}}{\sum q_{m,fuel} Q_{fuel,LHV}} \quad (7)$$

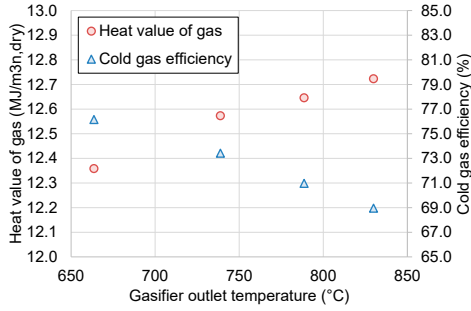


Fig. 12. Heat value of syngas and cold gas efficiency.

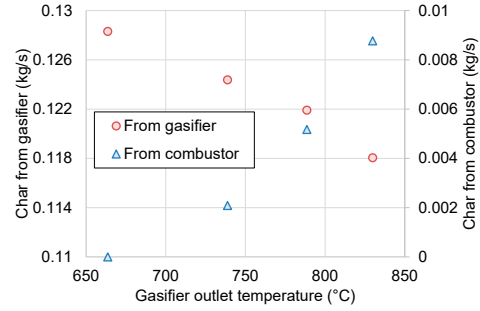


Fig. 13. Char flow rate from gasifier and combustor.

In direct air- or oxygen-blown gasification, the higher gasification temperature leads to lower heat value of gas. This is because the higher temperature is achieved by increasing the amount of oxygen, which increases the oxidation reactions and reduces the share of combustible gases. In indirect steam gasification, the higher temperature is achieved by increasing the fuel input to combustor. This increases the gas yield from gasification of char. Consequently, the heat value of gas is higher with a higher gasification temperature. As noted earlier, most of the produced syngas originates from the devolatilization. Consequently, the increase in the gas yield is relatively small and thus, the increment in the chemical heat of syngas is smaller than the increment of the fuel power. Consequently, the cold gas efficiency decreases with higher fuel input to combustor.

The unreacted char from gasifier is transferred to combustor with the other circulating solids. Based on these calculations, the amount of char from gasifier, which is burned in combustor is enough to sustain the gasification process. As the temperature increases, the water-gas reaction is enhanced and the char flow from gasifier decreases (Fig. 13). The char flow from combustor is considerably smaller as most of the char is burned. Increasing the temperature by higher fuel input to combustor increases the superficial gas velocity and decreases the residence time of char in the combustor. Consequently, this leads to higher char flow rate at the outlet of the combustor. At minimum temperature, when no fuel is fed to combustor, the char flow rate is almost zero, i.e. all the char from gasifier is burned in the combustor.

The modelled case does not represent any existing facility, thus, direct validation data is not available. Table 3 compares the main modelled parameters with the measurement data of indirect steam gasification from literature. The data of Technical University of Vienna (TUW) is from a 120 kW_{th} pilot plant using wood pellets (Pfeifer et al., 2009). The data of Güssing is from a 8 MW_{th} biomass gasifier (Specht, 2009). In both cases, the gasifier operated in BFB mode. There are some differences in the modelled syngas composition compared with measurements: higher H₂, lower CO, higher CO₂, and higher H₂O. The shares of hydrocarbons match the measurements well considering that the model does not simulate the heavier hydrocarbons (C₂H₆, C₃H₈), i.e. the modelled C₂H₄ can be interpreted to represent them as well. The differences can be partly due to different process conditions, e.g. higher fuel moisture, longer residence time, and BFB vs CFB mode. Other explanations are that the submodel for devolatilization predicted a too small CO/CO₂-ratio and/or the shift conversion rate was too high. The model parameters can be later adjusted when validation data is available. For now, the model results and the predicted effects with changing gasifier temperature appear to be quite reasonable.

Table 3. Modelled process data compared with measurement data from literature.

Parameter	Units	TUW	Güssing	Case A01	Case A02	Case A03	Case A04
Gasification temperature	(°C)	841	800...900	789	830	739	664
Combustion temperature	(°C)	920		884	922	836	767
Steam/fuel ratio	(-)	0.63	0.6...0.7	0.5	0.5	0.5	0.5
H ₂	(vol-%,dry)	39.1	38...40	43.8	43.5	44.3	44.9
CO	(vol-%,dry)	29.1	24...26	13.8	15.3	12.1	10.0
CO ₂	(vol-%,dry)	17.5	20...22	27.0	25.9	27.9	29.5
CH ₄	(vol-%,dry)	11.4	10...11	11.5	11.4	11.7	11.7
C ₂ H ₄	(vol-%,dry)	2.0	2.0...2.5	3.5	3.4	3.5	3.5
C ₂ H ₆ + C ₃ H ₈	(vol-%,dry)	0.9	0.5...1.0	n.m.	n.m.	n.m.	n.m.
H ₂ O	(vol-%,wet)	33.2	30...45	47.7	47.3	48.1	47.9
LHV	(MJ/m ³ n,dry)	13.7	12.9...13.6	12.65	12.72	12.57	12.36

CONCLUSIONS

The presented semi-empirical modelling method is an efficient tool for simulating indirect steam gasification in coupled gasifier/combustor CFB reactors. The model helps to analyse the different phenomena occurring in the reactors and to study the effects of process parameters on syngas composition and cold gas efficiency. Based on the simulations, the process can be operated without additional fuel to combustor. Increasing the fuel input to combustor increases the gasification temperature, which increases the heat value of the syngas slightly, but reduces the cold gas efficiency of the system. In the modelled cases, the limestone feed was small and the calcination and carbonation reactions did not contribute much to the syngas composition. In future, the model concept can be applied to study sorbent enhanced gasification, in which the CO₂ can be decreased in the gasifier by carbonation of lime.

NOTATION

C	molar concentration, mol/m ³	η_{CG}	Cold gas efficiency, -
$q_{m,fuel}$	fuel mass flow rate, kg/s		
$q_{v,gas}$	syngas flow rate, m ³ n/s	Subscripts	
$Q_{fuel,LHV}$	lower heat value of fuel, MJ/kg	<i>boud</i>	Boudouard reaction
$Q_{gas,LHV}$	lower heat value of gas, MJ/m ³ n	<i>i</i>	particle size group
r'''	reaction source term, mol/m ³ s	<i>shift</i>	shift conversion
T	temperature, K	<i>watg</i>	water-gas reaction

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